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Der Präsident des Europäischen Patentamts;
Im Auftrag

For the President of the European Patent Office

Le Président de l'Office européen des brevets
p.o.

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SUISSE

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If no title is shown please refer to the description.
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A process for the synthetic generation of methane

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Description

5

A process for the synthetic generation of methane

The invention relates to a process for the synthetic generation of methane from a feed gas mixture comprising
10 carbon monoxide, hydrogen and water vapour and optionally aromatic hydrocarbons.

It is the intention of the world-wide community to reduce the production of gaseous climate relevant components, such as
15 chlorofluorocarbons (CFC) and carbon dioxide. For the carbon dioxide the worldwide community committed itself to reduce the production within the next decade in the range of about 10 to 15%, depending on the country and its commitment to the Kyoto agreement.

20

So-called biofuels have the potential to influence the carbon dioxide balance tremendously on a midterm and long-term basis as being a relevant source for renewable primary products, such as wood or related biomass. The estimated cost are
25 expected in the range of 300 Swiss Francs per ton of CO₂. These costs are compared to other options taken in the transportation sector quite inexpensive.

Beside natural gas (methane) and hydrogen the biofuels are
30 categorized as potential alternative. On a short-term fuels like RME (bio diesel), ethyl alcohol and bio gas, are available; in mid- to long-term, bio -fuels will be produced according to the already theoretically well-known thermo-chemical processes which take advantage of a catalyst in order
35 to transform a synthesis gas into the desired bio fuel, such as methyl alcohol, synthetic gasoline or diesel, methane, or hydrogen.

A most promising fuel is methane that is supplied actually as fossil gas and which can be replaced under efficient use of synergies between the required infrastructure for both the fossil gas and the biogenic gas by the latter that can be produced by fermentation as well as by thermo-chemical processes.

Unfortunately, one of the most preferred renewable raw material, wood, cannot be transformed into bio gas by fermentation. Therefore, it is a crucial task to provide efficient alternative processes for the synthetic generation of methane from synthesis gas originating from wood gasification processes.

15 A process known in the art uses wood in a gasification reactor, such as FICFB (Fast Internally Circulation Fluidised Bed), which requires subsequently non-negligible efforts for the pretreatment of the raw synthesis gas in order to allow both effective and efficient synthesis of methane. As far as wood is considered as the renewable raw material, it has to be pointed out that the raw synthesis gas originating from the gasification step, is loaded with considerable amounts of aromatic hydrocarbons which are known in the prior art to have a negative impact in the subsequent process chain. Therefore, the removal of these aromatic hydrocarbons, such as benzene, toluene and naphthaline (BTN), is acknowledged as being a required process requisite as well as the removal of ammonia NH_3 and hydrogen sulfide H_2S . Unfortunately, these additional process requisites tend to increase the cost of the synthetic generation of methane from wood and reduce the overall efficiency.

For the reason given above, the crucial pre-requisite for an efficient exploitation of wood for the synthetic generation of methane is to find an optimized concert between the processes of gasification, raw synthesis gas purification and methane

generation. Of central importance are the properties of the catalyst required for the generation of methane.

Accordingly, it is the aim of the invention to provide a process which allows the synthetic generation of methane for a renewable raw material, explicitly including wood and related materials, in an economically and feasible manner.

This aim is achieved according to the invention by a process for the synthetic generation of methane from a feed gas mixture, i.e. a feed gas mixture originating from a biomass gasification process, comprising carbon monoxide, hydrogen and water vapour and optionally aromatic hydrocarbons; said process comprising the steps of:

- a) bringing the feed gas mixture in contact with a fluidized bed catalyst having catalyst particles which comprise as catalytic active component a metal and/or a metal compound or a mixture thereof;

under the circumstances of:

- b) an elevated temperature in the range of 250 to 500°C;
- c) a feed gas pressure in the range of 0.8 to 70 bar;
- d) a gas hourly space velocity of 1000 to 50000 h⁻¹; and
- e) a concentration of H₂/CO in the feed gas mixture in the range of 0.25 to 5.

The afore-mentioned process allows to catalytically convert hydrogen and carbon monoxide effectively in the fluidized bed. Using a fluidized bed catalytic reactor avoids a rapid deactivation of the catalyst material and therefore delivers a high activity of the catalytic active components in the process.

Both thermo-chemical reactions, the endothermic reformation of higher hydrocarbons, i.e. aromatic hydrocarbons, and the

exothermic methane generation, proceed simulataneously within the fluidized bed catalytic reactor, leading to an overall enhanced thermal efficiency of the conversion process.

- 5 With respect to the initial and continuing catalytic activity, excellent results have been achieved by using as catalytically active component nickel and/or a nickel compound, preferably a mixture of nickel and nickel monoxide, deposited on an ceramic carrier, such as Al_2O_3 , TiO_2 , SiO_2 or Y_2O_3 or mixtures thereof.
- 10 Additionally, the content of the catalytically active component may be in the range of 20 to 80 weight%, preferably 40 to 60 weight%, as compared to the weight of the catalyst particles. A suitable catalyst may comprise nickel and nickel oxide deposited on alumina (Al_2O_3) having a content of the
- 15 catalytically active component in the range of 50 weight% as calculated as neat nickel.

- In order to achieve both, a high mobility of the catalyst particles in the fluidized bed, and a sufficient reactive
- 20 surface area, the size of the catalyst particles may be in the range of 10 to 1000 μm , preferably in the range of 100 to 500 μm .

- With respect to the energy balance of the exothermic
- 25 generation of methane an advantageous feed gas composition is considered to have a positive impact thereupon when the feed gas mixture comprises aromatic hydrocarbons, such as benzene, toluene and naphthalene, in the range of less than 10 Vol%, preferably less than 5 vol%. Explicitly, this gas composition
- 30 covers broadly a synthesis gas that originates from wood gasification processes to which a major focus is laid upon due to the environmental demands.

An optimal yield with respect to the generation of methane as well as to a complete conversion of the higher hydrocarbons may require that the gas hourly space velocity (GHSV) is in the range of 2000 to 10000 h⁻¹, the temperature is in the range of 340 to 400°C and the gas pressure is in the range of 0.8 to 10 bar. With respect to the above-mentioned nickel/nickel oxide catalyst, the temperature may preferably be at about 350°C, and the GHSV at about 4200 h⁻¹.

- 10 In order to offer conditions in the fluidized bed catalyst that support both the catalytic reactions of reforming higher hydrocarbons and formation of methane and the regeneration of the catalyst particles, a mean residence time of the feed gas mixture in the fluidized bed catalyst may range from 0.1 to 5 sec., preferably 0.2 to 1 sec.

Again with respect to the yield of methane and to the reformation of higher hydrocarbons the content of H₂/CO in the feed gas mixture is in the range of 0.8 to 3, for example in the range of 1.5 as combined with the afore-mentioned nickel/nickel oxide catalyst.

Exemplarily embodiments of the inventive process are described in detail below without the intention to limit the invention to these actually preferred examples.

The measurements were taken with a plant as shown schematically below in Figure 1. The measurement campaign can be divided in three different sections:

30

Phase 1: The feed gas mixture is pretreated with a wet scrubberunit and an activated carbon filter (charcoal absorber) as proposed so far in the prior art which teaches the process as being unactable

without prior gas cleaning and removal of aromatic hydrocarbons;

5 Phase 2: The feed gas mixture is now only pretreated with the wet scrubber unit; charcoal absorber has been bypassed being aware of a possible catalyst deactivation reported so far in the prior art due to carbon which is deposited on the catalytic surfaces and by that blocks their reactivity (coking); and

10 Phase 3: The feed gas mixture is taken as of the outlet of the wood gasification plant without any pretreatment as disclosed in the prior art.

15 Especially for the intention of maintaining the activity and selectivity of the catalyst, unsaturated hydrocarbons (such as C_2H_4 , C_2H_2), light tar components (such as benzene, toluene, naphthalene, phenylacetylene, styrole, indene) and ammonia are known according to the prior art as being highly
20 disturbing these demands.

Figure 1 shows that the complete plant comprises an inlet at 1) for the feed gas mixture originating from a non-illustrated wood gasification plant, followed by a wet scrubber unit at
25 2), an activated carbon filter at 3), a gas pump and the methane generation unit COALA whereby the methane content is observed at the outlet of COALA at 4). The methane generation unit COALA comprises a desulphurisation unit in form of a ZnO fixed bed reactor, an inlet for the supply of water vapour, a
30 fluidized bed catalyst, a filter and a heat exchanger as seen in direction of the gas flow.

In phase 1, the washer unit for the removal of ammonia NH_3 and the activated carbon filter for the removal of higher
35 hydrocarbons are installed upstream the methane generation unit COALA. The gas composition during the first 40 hours of service is given in table 1; the first line giving the data

for the gas composition upstream the activated carbon filter, the second and third lines the respective gas compositions downstream the activated carbon filter after 3 and 40 h of service, resp.

5

Table 2 gives the gas compositions in phase 1 with respect to benzene, toluene, naphthalene and C8 at the probe locations 1), 2), 3) and 4). Almost all disturbing higher hydrocarbons are already removed at 3) by both the wet scrubber unit and the activated carbon filter which is the desired effect of gas purification being declared as a pre-requisite in the prior art for the efficient subsequent methane generation. The plant was operated continuously at a temperature of about 360°C, a water content of 0.25 as related to the dry gas volume and a gas volume flow rate of 0.6 m³/h.

15

During phase 2, the plant was operated without activated carbon filter (bypassing line named Phase 2+3 in Fig. 1) and the gas composition with respect to higher hydrocarbons is given in table 3. Surprisingly, a reduction of the content of higher hydrocarbons can be observed although the carbon filter unit has been bypassed.

20

During phase 3, the plant was operated without any pretreatment of the feed gas mixture according to the present invention. As already before in phases 1 and 2, for the fluidized bed catalyst 100 g catalyst particles having a size of 200 µm and comprising an even content of nickel and nickel oxide supported by alumina (Al₂O₃) whereby the content of the catalytically active nickel components is 50 weight% as calculated as neat nickel, was used. The gas composition for the higher hydrocarbons is given in table 4. At the entrance the average load of benzene, toluene, naphthalene and C8 was in the range of 13.6 g/Nm³ resp. 0.6 g/Nm³ resp. 0.8 g/Nm³ resp. 0.5 g/Nm³. At the outlet at 4) these higher hydrocarbons were reformed almost completely without the occurrence of any catalyst deactivation or loss in selectivity.

30

35

The comparison data given in table 5 is in so far self-explanatory, when the following definitions are observed:

5 The steam/dry gas ratio

$$D_v = \frac{n_{H_2O}}{n_{Gas\ tr.}} \quad [-]$$

is the molar ratio of water steam to all the other gas

- 10 components. The catalyst load is expressed as the ratio of the number of moles of gas brought into contact with a given mass of catalyst per hour.

$$KB = \frac{n_{Gas\ tr.} + n_{H_2O}}{M_{Kat}} \quad \left[\frac{\text{mol}}{\text{kg} \cdot \text{h}} \right]$$

15

originating from the generation of methane from synthesis gas, such as pure H_2/CO gas. The „steam to carbon ratio“

$$S/C = \frac{n_{H_2O}}{n_{Carbon}} \quad [-]$$

20

is the molar ratio of water steam to total amount of carbon in the gas and is known from the steam reforming of methane. All above mentioned key figures are used in a sense that clearly determined distinct gas compositions were used. Therefore,

- 25 they are not sufficient in order to describe the process of generating methane from a synthesis gas being achieved by a biomass gasification process. Additionally, the H/C ratio in both a dry product gas

30 $H/C_{tr} = \frac{n_H}{n_C}$

as well as in a humidified product gas

$$H/C_f = \frac{n_H + 2 \cdot n_{H_2O}}{n_C}$$

is a suitable parameter to characterize the process properly.

A further meaningful parameter is the conversion rate of

5 carbon monoxide

$$U_{CO} = \frac{n_{COin} - n_{COout}}{n_{COin}} \quad [-],$$

the selectivity to methane

10

$$S_{CH_4} = \frac{n_{CH_4in} - n_{CH_4out}}{(n_{COin} - n_{COout}) + x \cdot (n_{C_xH_yin} - n_{C_xH_yout})} \quad [-]$$

and the so called cold gas efficiency,

15

$$\eta_{KG} = \frac{\text{Heatingvalue}_{CH_4out}}{\text{Heatingvalue}_{in}}$$

which describes the heating value of the methane at the outlet in relation to the overall heating value of the gas at the inlet.

20

Surprisingly, the catalyst does not show any effect of deactivation and/or loss in selectivity which is indicated by the data given in the third and forth line. The third line stands for the results in phase 3a during the initial hours of operation. The results in the forth line stand for the gas compositions that have been achieved by continuously operating the plant over dozends of hours.

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Patent Claims

1. A process for the synthetic generation of methane from a feed gas mixture comprising carbon monoxide, hydrogen and water vapour and optionally aromatic hydrocarbons; said process comprising the steps of:

- a) bringing the feed gas mixture in contact with a fluidized bed catalyst having catalyst particles which comprise as catalytic active component a metal and/or a metal compound or a mixture thereof under the circumstances of:
- b) an elevated temperature in the range of 250 to 500°C;
- c) a feed gas pressure in the range of 0.8 to 70 bar;
- d) an gas hourly space velocity of 1000 to 50000 h⁻¹; and
- e) an concentration of H₂/CO in the initial gas mixture in the range of 0.25 to 5.

2. The process according to claim 1, characterized in that the catalytic active component is nickel and/or a nickel compound, preferably a mixture of nickel and nickel oxide, disposed on an ceramic carrier, such as Al₂O₃, TiO₂, SiO₂ or Y₂O₃ or mixtures thereof.

3. The process according to claim 2, characterized in that the content of the catalytically active component is in the range of 20 to 80 weight%, preferably 40 to 60 weight%, as compared to the weight of the catalyst particles.

4. The process according to claim 1, 2 or 3,
characterized in that
the size of the catalyst particles is in the range of 10 to
5 1000 μm , preferably in the range of 100 to 500 μm .
5. The process according to anyone of the preceding claims,
characterized in that
the feed gas mixture comprises aromatic hydrocarbons, such as
10 benzene, toluene and naphthalene, in the range of less than 10
Vol%, preferably less than 5 vol%.
6. The process according to anyone of the preceding claims,
characterized in that
15 the gas hourly space velocity is in the range of 2000 to 10000
 h^{-1} , the temperature is in the range of 340 to 400°C and the
gas pressure is in the range of 1 bar.
7. The process according to anyone of the preceding claims,
20 characterized in that
a mean residence time of the feed gas mixture in the fluidized
bed catalyst is in the range of 0.1 to 5 sec., preferably 0.2
to 1 sec.
- 25 8. The process according to anyone of the preceding claims,
characterized in that
the content of H_2/CO in the feed gas mixture is in the range of
0.8 to 2.

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Abstract

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The present invention discloses a process for the synthetic generation of methane from a feed gas mixture comprising carbon monoxide, hydrogen and water vapour and optionally aromatic hydrocarbons; said process comprising the steps of:

- 10 a) bringing the feed gas mixture in contact with a fluidized bed catalyst having catalyst particles which comprise as catalytic active component a metal and/or a metal compound or a mixture thereof under the circumstances of:
- b) an elevated temperature in the range of 250 to 500°C;
- 15 c) a feed gas pressure in the range of 0.8 to 70 bar;
- d) a gas hourly space velocity of 1000 to 50000 h⁻¹; and
- e) a concentration of H₂/CO in the initial gas mixture in the range of 0.25 to 5.

- 20 The afore-mentioned process allows to catalytically convert hydrogen and carbon monoxide effectively in a fluidized bed catalytic reactor which avoids a rapid deactivation of the catalyst material and therefore delivers a high activity of the catalytic active components in the process. Both thermo-
- 25 chemical reactions, the endothermic reformation of higher hydrocarbons, i.e. aromatic hydrocarbons, and the exothermic methane generation, proceed simulataneously within the fluidized bed catalytic reactor, leading to an overall enhanced thermal efficiency of the conversion process.

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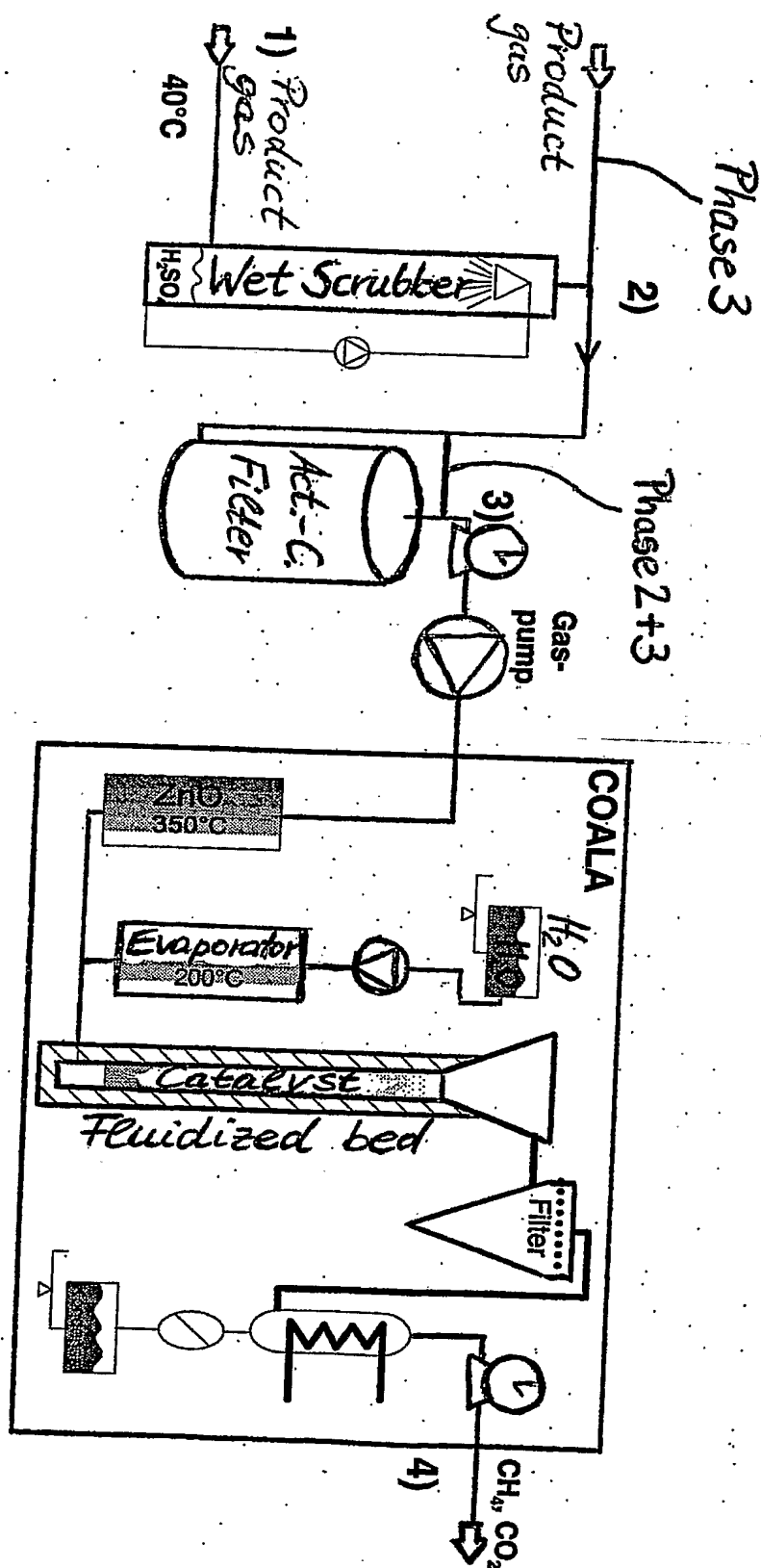


Fig. 1

Phase 1	CO2	C2H4	C2H6	C2H2	H2	O2	N2	CH4	CO
	%	%	%	%	%	%	%	%	%
Upstream Act.-C.	18.99	2.36	0.34	0	40.3	0.18	2.1	10.28	26.89
Downstream Act.-C. Filter (3h)	16.6	0.85	0.25	0	43.9	-	2.5	9.4	29.1
Downstream Act.-C. Filter (4h)	17.3	1.46	0.24	Estimals Peak	41.0	0.092	2.16	10.8	28.9

Table 1

Phase 1	Benzene	Toluene	Naphthalene	C8
	[mg/Nm ³]	[mg/Nm ³]	[mg/Nm ³]	[mg/Nm ³]
Inlet at 1)	21440	1068	1411	814
Inlet at 1)	14813	726	839	573
Inlet at 1)	7456	548	674	561
Product gas, downstream wet scrubber at 2				16
Inlet COALA, after wet scrubber & Act.-C. at 3	232	19		
Outlet COALA 4)	5	0		0
Outlet COALA 4)	2	0		0

Table 2

Phase 2		Benzene	Toluene	Naphthaline	C8
Inlet COALA after wet scrubber without Act-C Filter		[mg/Nm ³]	[mg/Nm ³]	[mg/Nm ³]	[mg/Nm ³]
Inlet COALA after wet scrubber, without Act-C Filter		9658	373	234	263
Inlet at 1)		13953	531	448	485
Inlet at 1)		21440	1068	1411	814
Inlet at 1)		14813	726	839	573

Table 3

Phase 3		Benzene	Toluene	Naphthaline	C8
Inlet COALA, without any pre-treatment		[mg/Nm ³]	[mg/Nm ³]	[mg/Nm ³]	[mg/Nm ³]
Inlet COALA, dito		17252	572	696	425
Inlet COALA, dito		13601	477	605	346
Inlet COALA, dito		11493	630	1013	670
Inlet COALA, dito		12152	571	723	510
Outlet COALA 4)		157	9	3	0
Outlet COALA 4)		136	9	1	0

Table 4

	V _{gas} Nl/min	V _{HzO} ml/min	T _{met} °C	D _v	KB	S/C	H/C _i	H/C _o	U _{co}	U _{BTM}	S _{CH4}	η _{kg}
BILANZ 1 Phase I	9.54	0.91	362.4	0.180	263.2	0.30	2.73	2.14	0.99	0.987	0.55	0.80
BILANZ 2 Phase II	10.42	0.95	359.8	0.177	262.5	0.27	2.62	2.07	0.97	0.990	0.64	0.84
BILANZ 3 Phase IIIa	10.80	0.94	357.0	0.164	299.7	0.27	2.54	2.03	0.96	0.992	0.65	0.84
BILANZ 4 Phase IIIb	7.44	1.04	360.60	0.229	217.6	0.35	2.72	2.03	0.98	0.992	0.64	0.85

Table 5